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Executive Summary for Full-Scale Dual-Alkali Demonstration at Louisville Gas and Electric Co. Preliminary Design and Cost Estimate

Louisville Gas and Electric Co, Ky.

Prepared for

Industrial Environmental Research Lab, Research Triangle Park, N.C.

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Office of Research and Development
Industrial Environmental Research Laboratory
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EXECUTIVE SUMMARY FOR FULL-SCALE DUAL-ALKALI DEMONSTRATION AT LOUISVILLE GAS AND ELECTRIC CO. - Preliminary Design and Cost Estimate

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16. ABSTRACT The report is the executive summary for the preliminary design of the dual-alkali system, designed by Combustion Equipment Associates, Inc./Arthur D. Little, Inc. and being installed to control SO2 emissions from Louisville Gas and Electric Company's Cane Run Unit No. 6 boiler. The project consists of four phases: I--preliminary design and cost estimates; II--engineering design, construction, and mechanical testing; III--startup and performance testing; and IV--1 year operation and testing. Developed as part of Phase I, the executive summary presents salient facts and conclusions from the Phase I report for use by upper management and the general public.

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by

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This report was prepared by Arthur D. Little, Inc.; however, the information and data contained in the report represent the work of many individuals from several organizations who have been involved in this project.

The principal participating organizations are Louisville Gas and Electric, Inc.; Combustion Equipment Associates, Inc.; and Arthur D. Little, Inc. The persons within each of these companies who were directly involved in the preparation of this report and the work performed during Phase I are listed below:

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In addition to the above, we would like to acknowledge the efforts and contributions from persons in other organizations. Norman Kaplan, the EPA Project Officer for this demonstration program, has made important technical contributions and has been instrumental in the management of the entire project. Mike Maxwell, the Director of Emissions/Effluent Technology at EPA's Industrial Environmental Research Laboratory, was responsible for overall planning and review for this program and has provided invaluable guidance and support. And Randall Rush of the Southern Company Services has made important contributions of a technical nature to the design of the system.

APPLICABLE CONVERSION FACTORS
ENGLISH TO METRIC UNITS

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 meter
1 ft^2	0.0929 $meters^2$
1 ft^3	0.0283 $meters^3$
1 grain	0.0648 gram
1 in.	2.54 centimeters
1 in^2	6.452 $centimeters^2$
1 in^3	16.39 $centimeters^3$
1 lb (avoir.)	0.4536 kilogram
1 ton (long)	1.0160 metric tons
1 ton (short)	0.9072 metric tons
1 gal.	3.7853 liters
1 Btu	252 calories

I. PURPOSE AND SCOPE

A. PURPOSE OF PROJECT

The project involves the full-scale application of a dual alkali flue gas desulfurization (FGD) process to a coal-fired utility boiler. The dual alkali system, developed jointly by Combustion Equipment Associates, Inc. (CEA) and Arthur D. Little, Inc. (ADL) is being installed by the Louisville Gas and Electric Company (LG&E) on Unit No. 6 at the Cane Run Station in Louisville, Kentucky. Unit No. 6 is an existing 300 megawatt (gross peak capacity), high sulfur, coal-fired boiler equipped with a high efficiency electrostatic precipitator. The dual alkali system is being installed to comply with requirements of the Jefferson County Air Pollution Control District, the Division of Air Pollution of the State of Kentucky, and Region IV of the U.S. Environmental Protection Agency (EPA)--i.e., removal of 85% of the SO_2 present in the flue gas entering the scrubber.

The system is being designed with the capability for controlling SO_2 emissions beyond this 85% removal requirement. When burning coal containing greater than 5% sulfur, the system will remove 95% of the SO_2 in the inlet flue gas. When burning coal containing less than 5% sulfur, the SO_2 emissions will be controlled to less than 200 ppm (dry basis). Since the system will follow an electrostatic precipitator, it is not designed for particulate removal. However, it is designed such that it will not increase the particulate loading in the flue gas.

EPA has selected the Cane Run dual alkali system as a demonstration unit for dual alkali technology and will participate on a cost-shared basis with LG&E for design, operation, testing, and reporting of the demonstration project. As a demonstration system, the purpose of the installation and operation is to establish:

- Overall performance--SO_2 removal, lime utilization, sodium makeup, regeneration of spent liquor, water balance, scale potential, materials of construction, waste cake properties, reliability, and availability; and

- Economics--capital investment and operating cost.

B. SCOPE OF WORK

LG&E will have overall responsibility for the design, construction, and operation of the system (including operation during the one-year test program). CEA, as a subcontractor to LG&E, will provide the engineering design for the dual alkali system and will supply all process equipment, instrumentation and control systems, and all process buildings. CEA is

also responsible for system startup and acceptance of the system as defined by the process guarantees. ADL will provide assistance to CEA in the process design and startup of the system and will assist LG&E in the operation during the one-year test program. In addition, ADL is also responsible for reporting progress during each phase of the demonstration program.

The work under the demonstration program is divided into four phases:

- Phase I - preliminary design and cost estimate;
- Phase II - engineering, design, construction, and mechanical testing;
- Phase III - startup and performance testing; and
- Phase IV - one year of operation and testing.

The overall project schedule covering all phases of the dual alkali demonstration project is shown in Figure I-1. The overall project, including the one-year test program, is scheduled for 40 months (with an additional one month for completion of the final draft report). In order to expedite the overall project, Phases I and II were scheduled to begin simultaneously and run concurrently. Phase I (preliminary design) was scheduled for five months, including preparation of the draft report. Phase II (engineering, design, and construction) is scheduled for 24 months. A three-month period has been allowed for Phase III (startup and acceptance testing). The one-year test program (Phase IV) is scheduled to start in late 1978, approximately 27 months after the initiation of the project.

This summary (EPA-600/7-78-010A) covers work performed under Phase I of the project as reported in the Project Manual. The Project Manual (EPA-600/7-78-010) contains information generated in preparation of the preliminary design in sufficient detail to convey the total concept of the proposed plant and to provide a complete basis for the cost estimates. Included in the Project Manual are: a general description of the process technology and its adaptation to Cane Run Unit No. 6; detailed material and energy balances; overall utility requirements; plot plans and the general arrangement of equipment; specification of major equipment; an instrument list; piping and instrumentation diagrams; a discussion of offsites, and laboratory and shop capabilities; and the project schedule. A detailed estimate of capital and operating costs for the demonstration plant has also been prepared based on the process design. These costs are broken down into the components used in preparing the estimate.

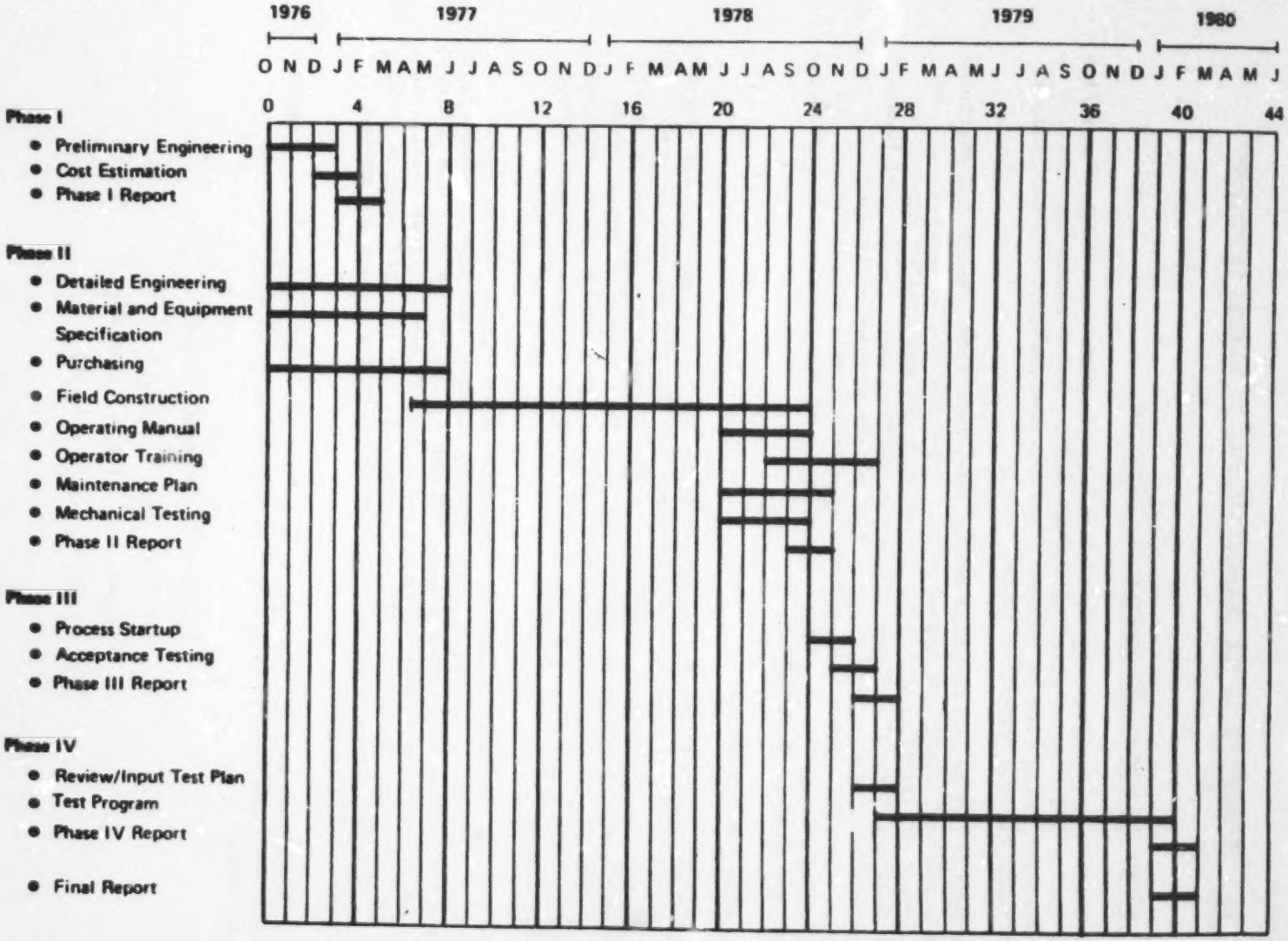

FIGURE I-1. DUAL ALKALI DEMONSTRATION OVERALL PROJECT SCHEDULE

II. DESCRIPTION OF THE CEA/ADL DUAL ALKALI PROCESS TECHNOLOGY

A. SYSTEM CHEMISTRY AND PROCESS CONFIGURATION

The CEA/ADL dual alkali SO_2 control process involves scrubbing of the flue gas with a solution of alkaline sodium salts. Spent scrubbing solution is regenerated using lime to produce a solid waste consisting principally of a mixture of calcium-sulfur salts.

The system can be conveniently broken down into three processing sections: gas scrubbing; absorbent regeneration; and waste solids dewatering. The equipment utilization and operation of each process section depends on specific requirements of the particular application. Figure II-1 shows a generalized flow schematic of a dual alkali process as applied to a boiler equipped with a high efficiency electrostatic precipitator. The following discussion provides a description of the basic system configuration and process flow scheme.

1. Flue Gas Scrubbing

In the generalized system, shown in Figure II-1, flue gas from the electrostatic precipitator (or induced draft fan for the boiler) is forced by a booster fan through an absorber. In the absorber the gas passes upward through a set of sprays to quench the gas, then through a set of sieve trays for SO_2 removal, and finally through a demister to remove entrained liquor. The clean flue gas leaving the tower is finally reheated before being discharged to the stack.

Regenerated absorbent solution, containing a mixture of sodium salts (sodium hydroxide, sodium sulfite, sodium sulfate, and some carbonate), is mixed with scrubber recycle liquor and fed to the top tray of the absorber. The solution flows countercurrent to the gas through the tray system and is collected at the bottom of the absorber in the internal recycle tank. The collected liquor supplies solution for both the quench sprays and recycle to the trays.

The absorption of SO_2 and some oxygen from the flue gas produces a spent sodium sulfite/bisulfite, sulfate liquor as shown in the following reactions:

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O$$

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2\uparrow$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3$$

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4$$

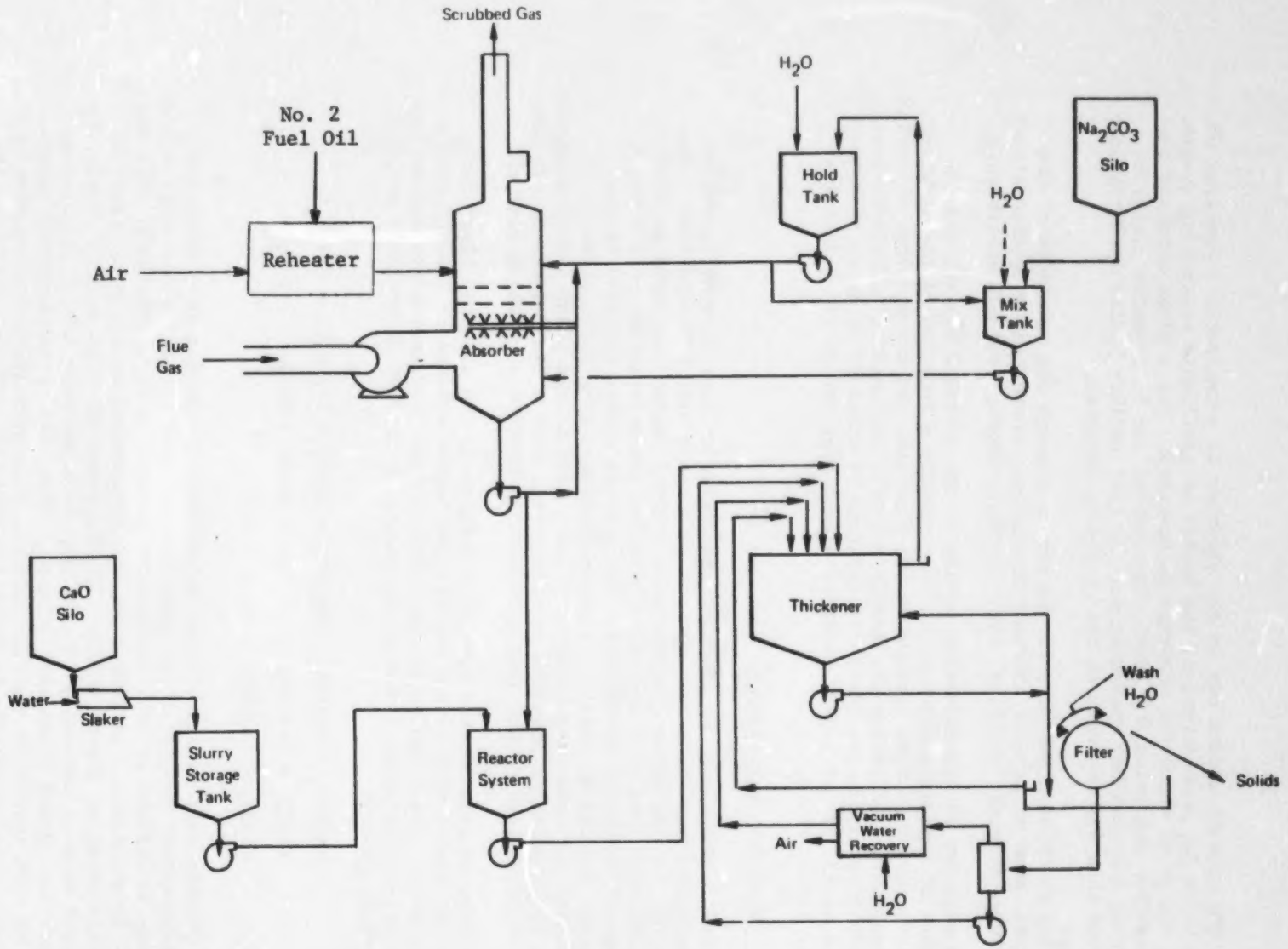

FIGURE II–1 DUAL ALKALI PROCESS FLOW DIAGRAM

The SO_2 removal efficiency in the absorber is principally a function of the inlet SO_2 concentration, the number of gas/liquid contacting stages, and the pH at which the absorber is operated. For a given scrubber configuration and inlet SO_2 level, SO_2 removal can be adjusted simply by varying the scrubber operating pH. In most utility boiler applications, better than 95% SO_2 removal can be easily achieved.

During absorption and to a lesser extent through the remainder of the system, some sulfite is oxidized to sulfate, converting an active alkali (Na_2SO_3) into an "inactive" salt (Na_2SO_4)--Na_2SO_4 does not absorb SO_2.

The majority of the oxidation occurs in the scrubber system; and the amount of oxidation experienced is generally a function of the scrubber configuration, oxygen content of the flue gas, and the scrubber operating temperature. At excess air levels normally encountered in utility power plant operations, the level of oxidation is expected to be on the order of 5 to 10% of the sulfur dioxide removed (for medium and high sulfur coal applications).

2. Absorbent Regeneration

A bleed stream from the absorber recirculation loop is sent to the regeneration reactor system where it is reacted with hydrated lime. The CEA/ADL reactor system incorporates a novel design developed to produce solids with good settling and filtration characteristics over a broad range of operating conditions. The reactor system consists of two reactor vessels in series: a short-residence time first stage (3-15 minutes) followed by a longer residence time second stage (20-40 minutes). The process can be operated in conjunction with a lime slaker or can use purchased hydrated lime (e.g., commercial hydrate or carbide sludge).

The lime neutralizes the bisulfite acidity in the scrubber bleed and further reacts with sodium sulfite and sodium sulfate to produce sodium hydroxide. These reactions precipitate mixed calcium sulfite and sulfate solids, resulting in a slurry containing up to 5 wt % insoluble solids, as shown below:

$$2NaHSO_3 + Ca(OH)_2 \rightarrow Na_2SO_3 + CaSO_3 \cdot 1/2\ H_2O\downarrow + 3/2\ H_2O$$

$$Na_2SO_3 + Ca(OH)_2 + 1/2\ H_2O \rightarrow 2NaOH + CaSO_3 \cdot 1/2\ H_2O\downarrow$$

$$Na_2SO_4 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_4\downarrow$$

The CEA/ADL dual alkali process is designed to operate in a relatively "concentrated" active alkali mode (roughly 0.5M active Na^+ -- where active sodium is defined as that sodium attributable to OH^-, $CO_3^=$, HCO_3^-, $SO_3^=$, and HSO_3^- in solution). In this mode, sulfate removal cannot be effected by precipitation of gypsum ($CaSO_4 \cdot 2H_2O$), since the high sulfite levels prevent soluble calcium concentrations from reaching levels required to exceed the gypsum solubility product. Thus, the system operates unsaturated with respect to calcium sulfate, thus minimizing scale potential. However, calcium sulfate is coprecipitated with calcium sulfite in the regeneration reactor, which allows the system to keep up with absorbent oxidation.

3. Solid/Liquid Separation and Solids Dewatering

Slurry from the regeneration reactor system is fed to a thickener where the slurry is concentrated to 20-45% solids. Clarified liquor from the thickener is collected in a hold tank from which it is returned to the absorber. The thickened slurry is sent to a rotary drum vacuum filter, where the solids are filtered to a cake containing 55-70% solids. The filter cake is the only waste material generated by the process. There are no other solid or liquor purges from the system.

The high solids content of the filter cake and the excellent handling properties of the material are a direct result of the controlled conditions for crystallization in the reactor system. The material is much like a moist silty to sandy soil and is easy to manage in solids handling and transport equipment. If further chemical treatment is required, these excellent physical properties should prove to be an advantage.

On the filter the cake is washed using a series of water spray banks. This wash removes a large fraction (approximately 90%) of the occluded soluble salts from the cake and returns these salts along with the filtrate to the system, thereby reducing sodium losses and minimizing sodium carbonate makeup.

The exact chemical composition of the solids will depend somewhat upon fly ash loading (if any), the chemical composition of the flue gas and fly ash, and the degree of oxidation of sulfite to sulfate encountered in the system. The following general chemical composition (ash-free, dry cake basis) is typical for a high sulfur coal application:

$CaSO_3 \cdot 1/2\ H_2O$	= 80-85 wt %
$CaSO_4 \cdot 1/2\ H_2O$	= 10-15 wt %
$Na_2SO_4 + Na_2SO_3 + NaCl$	= 1-3 wt %
$CaCO_3$ + Inerts	= 5-10 wt %

The mix of sodium salts in the cake is importantly dependent upon the chloride content of the coal. Essentially all of the chloride in the coal is released to the flue gas as HCl gas, and it is efficiently removed in the absorber. The chloride is purged from the system in the filter cake (as NaCl) at the rate at which it is absorbed. Thus, the higher the chloride content of the coal, the higher the fraction of sodium salts as sodium chloride.

Makeup sodium carbonate solution is fed to the system to replace sodium value lost in the filter cake. The sodium carbonate is not intended for use as a softener, since soluble calcium concentrations in the regenerated liquor runs less than 100 ppm. The quantity of sodium carbonate required, therefore, is quite small, and the sodium carbonate costs represent a negligible element in the overall cost of operating the system.

B. POLLUTION CONTROL CAPABILITIES

1. SO_2 Control

A sodium-based dual alkali process, operating in the concentrated active sodium mode, is capable of SO_2 removal efficiencies in excess of 95% over almost any range of inlet SO_2 concentrations encountered in coal-fired utility boiler applications. In most high sulfur coal applications, removal efficiencies approaching 99% can be achieved on a continuous basis in low-energy, tray-type absorbers. SO_2 removal efficiencies can be easily varied by adjusting the operating pH of the absorber with little or no affect on the overall lime stoichiometry or the sodium makeup requirement.

The high SO_2 removal capability of this process, when used in conjunction with a boiler equipped with adequate particulate control, allows the option of removing virtually all of the SO_2 from the flue gas treated in the scrubber. Thus, there is the option of bypassing hot, untreated gas to provide part or all of the reheat while still meeting the overall plant SO_2 emission regulations in the combined treated and untreated flue gas. In such a system the scrubber size can be reduced, since not all flue gas is treated. Also, reheat requirements are reduced or eliminated. In the system for Cane Run Unit No. 6, bypass for reheat is not provided. All of the flue gas will be treated with reheat provided by injection of hot gas from the combustion of No. 2 fuel oil.

2. Particulate Control

While particulate removal capability is not incorporated in the Cane Run system, it can be accommodated in the process by appropriate selection of the type of scrubber system to be used. If particulate removal capability is to be incorporated in the system, then a higher energy particulate removal device, such as a venturi scrubber, would be required. This can be used alone for both SO_2 and particulate control; or, if additional SO_2 removal is required (>95%), a venturi can be used in combination with one or more trays at a small incremental cost. Particulate removal down to 0.02 grains/scfd or lower can be accomplished using venturi scrubbers at moderate pressure drops (about 15 inches WG).

Where particulate control is already provided ahead of the scrubber system, the SO_2 scrubbing system will not result in any increase in particulate emissions. Because a solution is used for SO_2 absorption rather than a slurry, any entrainment of scrubbing solution in the gas can be efficiently and reliably removed using standard mist eliminators without wash water prior to exhausting the gas.

3. Chloride Control

Most of the chlorides in coal (>90%) is volatized and appears in the flue gas as HCL. Any aqueous-based scrubbing system would be highly effective in absorption of HCl (as well as HF) from the flue gas. As a result, chloride concentrations will build in the closed liquor loop to levels

such that the rate at which chloride is discharged from the system in the washed filter cake will equal the rate at which chloride enters the system with the flue gas. The CAE/ADL dual alkali system has been successfully operated for extended periods with steady-state levels of chloride in the closed liquor loop as high as 11,000 ppm (consistent with 0.05-0.10 wt % chloride in the coal).

III. DUAL ALKALI SYSTEM APPLICATION TO CANE RUN UNIT NO. 6

A. SYSTEM DESIGN

1. Design Basis

Cane Run Unit No. 6 is a 280 megawatt, pulverized coal-fired boiler. It is equipped with a high efficiency electrostatic precipitator capable of 99.4% particulate removal. The unit operates from a maximum load of 60 megawatts during peak hours to a minimum load of 60 megawatts during off-peak hours (20% of gross peak capacity). The annual average load is equivalent to approximately 180 megawatts (about 60% of the gross peak capacity). The system is designed to operate over the full range of anticipated boiler loads.

Coal for Unit No. 6 is received from a number of sources. The average sulfur content of the coal on a dry basis is 4.8% and varies from 3.5% to 6.3%. The average chloride content of the coal is 0.04% and varies from 0.03% to 0.06%. The average 4.8% sulfur content and 11,000 Btu/lb corresponds to an average SO_2 emission of about seven times that allowed by the present Federal New Source Performance Standards (1.2 lbs of SO_2/ MM Btu).

At present, Cane Run Unit No. 6 is operating in accordance with all applicable federal, state, and local regulations covering air emissions, water discharges, and operator safety and health. Construction permits have been issued by the Jefferson County Air Pollution Control District for the construction of an SO_2 removal system for Cane Run Unit No. 6. The permit complies with the two enforcement orders issued by the EPA (November 5, 1975) and the Jefferson County Air Pollution Control District (December 10, 1975).

The dual alkali system for Cane Run Unit No. 6 is designed for SO_2 removal only. The design basis for the system is summarized in Table III-1. Design conditions correspond to coal containing 5% sulfur and 0.04% chloride, and having a heating value of 11,000 Btu/lb. The design gas capacity of 3,372,000 lbs flue gas/hour is equivalent to the boiler peak load capacity of 300 megawatts. The dual alkali system will meet all applicable federal, state, and local pollution control and safety regulations. The maximum SO_2 concentration in the clean gas will be 200 ppm (for coal containing up to 5% sulfur), well below requirements of the Federal New Source Performance Standards.

2. System Description

The system configuration and operation will differ slightly from the generalized process discussed previously. The principal difference is that the system will utilize locally available carbide lime, a calcium hydroxide byproduct from acetylene production, rather than slaked quicklime.

TABLE III-1

DUAL ALKALI PROCESS DESIGN BASIS

Coal (Dry Basis):	
Sulfur	5.0% S
Chloride	0.04% Cl
Heat Content	11,000 Btu/lb
Inlet Gas:	
Flow Rate (Volumetric) (Weight)	1,065,000 acfm 3,372,000 lb/hr
Temperature	300°F
SO_2	3471 ppm
O_2	5.7%
Particulate	$\leq$0.10 lb/10^6 Btu
Outlet Gas:	
SO_2	$\leq$200 ppm ($\sim$0.45 lb/10^6 Btu)
Particulate	$\leq$0.10 lb/10^6 Btu

The system design is modular in nature, with spare capacity provided both as excess capacity within modules as well as spare modules and equipment where appropriate. Redundant instrumentation has also been provided for critical control operations.

a. Absorption System

The sulfur dioxide system will be installed between the existing induced draft fans and stack. Hot flue gas will be drawn from the existing ducting at the discharge of the two induced draft fans downstream of the electrostatic precipitator and fed to the dual alkali scrubber system through two booster fans. There are two absorber modules, with one booster fan for each module. A common duct connects the two inlet ducts to the booster fans (not shown in Drawing No. 040044-1-1, Rev. G of the Project Manual, EPA-600/7-78-010).

In each absorber the hot flue gas is cooled by sprays of absorber solution directed at the underside of the bottom tray. These sprays keep the underside of the tray and the bottom of the absorber free of buildup of fly ash solids. The cooled gas then passes through a set of two trays, where SO_2 is removed, then through a chevron-type mist eliminator. The saturated gas from the demister is heated 50 F° (to 175°F) by hot flue gas from a reheater fired with No. 2 oil, and the reheated gas is then returned to the existing duct entering the chimney.

Each absorber is sized to handle 60% of the design gas flow rate, and the system can be turned down to 20% of the design flow rate. At levels less than 60% of the design capacity the system can be operated with one absorber module (by using a common duct and shutting down one absorber module) or with two absorber modules.

Dampers are provided in the existing ducting and the absorber inlet ducts to allow bypassing of flue gas around the scrubber modules. These dampers are interlocked to enable bringing scrubbers on- or off-line without interruption of the boiler operation. The common ducting and damper systems allow each scrubber to be independently isolated for maintenance while the other scrubber continues in operation.

Regenerated absorbent liquor is fed to the top tray along with recirculated liquor from the absorber recycle tank for pH control. The feed forward of regenerated liquor from the thickener hold tank to the scrubber trays corresponds to an L/G of 4.0 gals./1,000 acf (saturated) at design conditions. The total recirculation rate for each absorber (sprays plus trays) corresponds to an L/G of 5.7 gals./1,000 acf.

b. Regeneration System

There are two identical reactor trains in the system. Each train consists of a primary reactor, a secondary reactor, and a reactor pump. The spent liquors from the absorbers are fed to the primary reactors of each two-stage reactor system along with slurried carbide lime from the carbide

lime storage tank for the dual alkali system (day tank). Each primary reactor overflows by gravity into a secondary reactor, where the reaction between lime and sodium salts is completed. The reactor slurry is then pumped to the solid/liquid separation section.

Normally, the number of reactor trains in use is the same as the number of operating scrubber modules, with each reactor train regenerating the spent liquor from the corresponding scrubber module. However, piping is provided such that each module can be operated on liquor from the corresponding scrubber or liquor from both the scrubbers. And the reactors are sized so that for short periods a reactor train can handle the total liquor from both the absorbers operating at design conditions. Thus, maintenance can be provided to one reactor train while the system is operated with the other.

The carbide lime slurry used for regeneration of the spent absorber solution is received at the plant as a 30% solids slurry and stored in a large tank for use in the three FGD systems. (Lime slurry scrubbing processes have been installed on Unit 4, a 180-megawatt boiler, and on Unit 5, a 170-megawatt boiler.) From this main storage the slurry is pumped to a day tank for each system. Slurry from the dual alkali system day tank is pumped to the primary reactors as required.

c. Solids/Liquid Separation

The reactor effluent streams are fed to a thickener. Clarified liquor overflows to the thickener hold tank from which the regenerated solution is pumped to the absorbers as required. As discussed before, the feed rate to the absorbers is determined by the pH of the absorber liquor. The total volume in the system is maintained by controlling the liquid level in the thickener hold tank using process makeup water.

The thickener underflow slurry controlled at about 25% solids is pumped to the filter system where solids separation is completed. The filter cake is washed to recover the sodium salts in the liquor. Combined filtrate and wash water is returned to the thickener.

There are three filters. Each filter is rated to handle 50% of the total solids produced at the design conditions, and each can be operated independently. The number of filters in operation is determined by the quantity of solids accumulated in the thickener, which is reflected in the solids concentration in the underflow slurry.

Soda ash is added to the system to make up for the sodium salts lost in the washed waste filter cake. Dry, dense soda ash is received at the plant and stored in the soda ash silo from which it is continuously weighed and fed to the soda ash solution tank. The soda ash solution is prepared using thickener hold tank liquor.

d. Provisions for Spills and Leaks

Filter cake is the only product of the dual alkali process. The system is operated in a closed loop, and there is no other liquid or solid purge from the system. In practice, though, some liquor can be lost through pump seals and during flushing of pipes and pumps after individual pumps are shut down or after the system is shut down. The concentration of the liquor is generally low compared to the concentration in the process liquor because of dilution with seal or flush water. However, to prevent water pollution and to reduce sodium loss from the system, this solution is collected in sump tanks and is returned to the thickener.

e. Waste Disposal

A long-range plan for the disposal of the dual alkali filter cake is now being developed as a part of an overall disposal plan for all FGD sludges produced at the Cane Run Station. The plan, in its preliminary form, has been submitted to the Jefferson County Air Pollution Control Board.

As currently perceived, the disposal operation will involve mixing dual alkali filter cake with thickener underflow from the direct lime scrubbing systems on Cane Run Unit Nos. 4 and 5, producing a material of approximately 40% solids. The combined wastes will then be transported to an onsite disposal pond. Prior to being placed in the pond, the combined sludge will be admixed with dry fly ash collected from the electrostatic precipitators and separately conveyed to the disposal area. Additives, such as lime, may also be used to promote hardening of the material.

Studies are now underway at the University of Louisville to develop adequate physical properties data to allow for the design of appropriate handling, mixing, and transport facilities. It is hoped that the dual alkali filter cake and direct lime thickener underflows can be handled as a thick slurry capable of being piped to the disposal area.

Since this combined disposal system is still in planning, the necessary handling and transport equipment will not be operational at the time the dual alkali system starts up. Thus, in the interim between the startup of the dual alkali system and installation of the equipment for the combined waste disposal operation, the filter cake will be conveyed to a feed hopper where it will be loaded into trucks for transport to the disposal pond.

f. Offsites

The offsites generally required for the dual alkali system include: services for electrical supply, water supply, and instrument air; oil, steam, or hot water if the wet flue gas is to be reheated; raw materials receiving and storage facilities; a wet chemicals analytical laboratory; and appropriate shop facilities for repair and maintenance of machinery and instruments. Except for electrical service, all of these offsites, including lime receiving and storage facilities, now exist at Cane Run Station and are available. An electrical substation including appropriate step-down transformers will be installed for the dual alkali system.

3. Materials of Construction

The system is designed with appropriate corrosion resistance where required using stainless steel (316 or 317) or linings (polyester or rubber). The expected chloride levels in the process liquor range from 10,000 ppm to 15,000 ppm, but levels can vary from as low as a few thousand ppm to almost 20,000 ppm depending upon the chloride content of the coal and the degree of cake washing. Liquor pH's range from about 5.0 in the absorber loop to greater than 12.0 in the reactor and dewatering systems.

With the exception of the primary reactors, all tanks and vessel linings in contact with process liquor are lined with flake-reinforced polyester. The primary reactor is constructed of 316L stainless steel; the filtration equipment is both 316L stainless steel and fiberglass; and the absorber trays are 317L stainless steel.

All pumps and agitators in contact with process liquor are rubber-lined. Process liquor piping is FRP. Hot flue gas ducting is carbon steel, and the booster fan housing and blades are A441 steel. Saturated flue gas ducting is polyester-lined between the absorbers and reheater section and is 317L stainless steel between the reheater and stack.

4. Instrumentation

The system is designed for minimal operator interface consistent with safe and reliable operation. Feed rates for all raw materials and the flows of all principal process streams are automatically controlled according to process operating conditions (tank liquid levels, process flows, and stream compositions). Non-critical internal process flows are preset and adjusted intermittently as dictated by process requirements.

All remote controls are located in a centralized control room from which the system can be started up, operated, and shut down. The control room is furnished with appropriate controllers, indicators, recorders, alarms, and other necessary instrumentation for the safe and convenient operation of the system.

Redundancy has been provided in critical control elements to ensure smooth operation and minimize downtime. Instrumentation in addition to that required to operate the process has also been provided to permit accurate calculation of process material and energy balances. In particular, instruments are included for continuous monitoring of inlet and outlet SO_2 concentrations, the measurement of all chemicals and water entering the system, and the weighing of all filter cake leaving the system.

B. OPERATING REQUIREMENTS

System operating requirements at design conditions are summarized in Table III-2. The power required to operate the process (exclusive of reheat) will be approximately 1.0% of the peak power generation for Unit No. 6. Of this, approximately 60% is required for the booster fans, 10% for reheater fans, and 30% for the remainder of the system. Including oil for reheat, the total energy requirement for the system amounts to about 2.7% of the peak power generation.

The lime and soda ash makeup requirements correspond to feed stoichiometries of 1.00 mols $Ca(OH)_2$ per mol of SO_2 removed and 0.022 mols Na_2CO_3 per mol of SO_2 removed. These feed requirements include the alkali needed for chloride removal. The total wet cake produced is equivalent to 30 lbs of moist filter cake per 100 lbs of coal (5% sulfur) fired.

C. GUARANTEES

1. Sulfur Dioxide Emission

The system shall provide such control that emissions from the stack shall be no greater than 200 ppm SO_2 dry basis without additional air dilution when burning coal containing less than 5% sulfur. When burning coal containing 5% sulfur or greater, the system shall be designed to remove at least 95% of the sulfur dioxide in the inlet flue gas.

2. Particulate Matter Emission

In addition to meeting applicable regulations, the system shall also be designed to meet Federal New Source Performance Standards for emissions of particulate matter under all conditions of boiler operation. The dual alkali system will not add any particulate matter to the emissions of particulate matter that is received by the system from the LG&E Cane Run Unit No. 6 electrostatic precipitator.

3. Lime Consumption

The consumption of lime in the system will not exceed 1.05 mols CaO per mol of SO_2 removed from the flue gas.

4. Sodium Carbonate Makeup

Soda ash makeup will not exceed 0.045 mols of Na_2CO_3 per mol of SO_2 removed from the flue gas, provided that the chloride content of the coal burned averages 0.06% or less. If the average chloride content of the coal is above 0.06%, then additional sodium carbonate consumption will be allowed at the rate of 1/2 mol Na_2CO_3 for each mol of chloride in the flue gas resulting from chloride in excess of 0.06% in the coal. The seller as part of the guarantee shall perform the necessary research and design to reduce the makeup requirements of Na_2CO_3 from the guarantee point to a level of approaching minimal makeup.

TABLE III-2

PROCESS OPERATING REQUIREMENTS AT DESIGN CONDITIONS

Basis: Coal - 5.0% sulfur
- 0.04% chloride
- 11,000 Btu/lb

SO_2 Removal - 94.4%
Full Load (300 megawatts)

	Consumption Rate	Equivalent % of Boiler Capacity
Energy Requirements		
Power	3.1 megawatts	1.03
Fuel Oil (for reheat)	48×10^6 Btu/hr	1.68
Total		2.7
	Consumption Rate	**Lbs/lb Coal Fired**
Makeup Materials		
Water	325 gpm	--
Lime (as $Ca(OH)_2$)	460 lbs/min.	0.111
Soda Ash	13.7 lbs/min.	0.003
Cake Production		
Dry Basis	804 lbs/min.	0.194
Wet Basis	1,246 lbs/min.	0.300

5. Power Consumption

At the peak operating rate (300 megawatts) the system will consume a maximum of 1.1% of the power generated by the unit.

6. Waste Solids Properties

The waste produced by the vacuum filter will contain a minimum of 55% insoluble solids.

7. SO_2 System Availability

The system will have an availability (as defined by the Edison Electric Institute for power plant equipment) of at least 90% for one year. Thus, the system will be available for operation at least 90% of the calendar time.

IV. CAPITAL AND OPERATING COSTS — DUAL ALKALI SYSTEMS

Estimates of the capital and operating costs for the demonstration dual alkali system to be installed on Unit No. 6 at LG&E's Cane Run Station have been prepared based upon the preliminary engineering in Phase I. Using these estimates, generalized costs have also been developed for application of dual alkali systems to new 500 and 1,000 megawatt boilers.

A. 300 MEGAWATT RETROFIT SYSTEM FOR LG&E CANE RUN UNIT NO. 6

The capital investment and operating cost estimates for the dual alkali demonstration system are summarized in Table IV-1. Capital costs are broken down into total materials, erection, engineering, spare parts, and working capital. The costs are as incurred and therefore are roughly equivalent to September 1977 dollars. The estimated capital investment for the system is $17,379,000, which corresponds to about $58.0/kilowatt (based on gross peak capacity).

This capital investment represents the total installed cost for the system exclusive of land and interest during construction. Since this is a retrofit system installed alongside two existing direct lime scrubbing systems, some cost savings are realized in common offsite facilities. These include the additional capacity requirements for service water and plant air, storage for reheater oil, and the switch gear building--none of which are included in the total capital. Also, the capital costs for the lime handling, transport, and storage facilities for all three systems have been included as a part of the delivered cost of lime in the operating costs.

Estimated annual operating costs for the demonstration system are shown in Table IV-1, both in 1979 dollars (the first full year of operation) and in corresponding 1976 dollars. Costs in 1976 dollars are shown since these represent more well-defined costs using recent unit cost data and should be directly comparable with published cost data for FGD systems. These estimates have been prepared on present and projected (1979) costs for labor, utilities, and raw materials, and incorporate LG&E's cost structure applicable to the plant for overhead rates, interest, and depreciation.

Operating costs are based on an average sulfur content in the coal of 3.8%, with an average boiler load of 60% and an average sulfur dioxide removal efficiency of 94.2%. The operating costs are estimated both with and without operation of the reheat system. (When included, reheat is assumed to be 50 F°.) The sludge disposal costs are based on onsite solids disposal after mixing with ash and lime.

The total annual operating cost without reheat is estimated to be $3.6 million in 1976 and $4.3 million in 1979. The corresponding costs with reheat are about 15% higher. The 1976 operating costs range from 2.3 mills per kwh (23.0¢ per million Btu) without reheat system to 2.7 mills per kwh (26.7¢ per million Btu) with 50 F° reheat. The cost estimates for the year 1979 are about 20% higher.

TABLE IV-1

SUMMARY OF CAPITAL AND OPERATING COST ESTIMATES FOR THE DUAL ALKALI SYSTEM AT THE LG&E CANE RUN UNIT NO. 6

Basis: 300 megawatt (gross peak load)
60% annual load
9,960 Btu/kwh
3.8% S in coal
94.2% SO_2 removal

CAPITAL COST ESTIMATES

Total Materials	$ 11,162,000
Erection	3,307,000
Total Engineering	2,478,000
Spare Parts	232,000
Total Fixed Plant	17,179,000
Working Capital	200,000
Total Capital	17,379,000
Capital Cost Equivalents	
$/kw - Based on 300 Mw Gross Peak Load	$ 57.9
$/kw - Based on 277 Mw Net Peak Load	$ 62.7

OPERATING COST ESTIMATES

	1979 Dollars		1976 Dollars	
	With Reheat (50F°)	No Reheat	With Reheat (50F°)	No Reheat
Annual Operating Costs				
Direct Costs	$3,019,100	$2,243,900	$2,344,200	$1,767,300
Indirect Costs	2,082,300	2,082,300	1,841,200	1,841,200
Total	5,101,400	4,326,200	4,185,400	3,608,500
Operating Cost Equivalents				
Mills/kwh	3.24	2.74	2.65	2.29
¢/10^6 Btu	32.5	27.6	26.7	23.0
$/ton S Removed	218	185	179	154

B. GENERALIZED CAPITAL AND OPERATING COSTS

Generalized costs have been prepared for application of concentrated mode, lime-based dual alkali systems for SO_2 control on new 500 and 1,000 megawatt boilers. Three cases have been considered: (1) a new 500 megawatt boiler with SO_2 removal necessary to meet current Federal New Source Performance Standards (NSPS) - 78.1% SO_2 removal; (2) a new 500 megawatt boiler with 90% SO_2 removal; and (3) a new 1,000 megawatt boiler with SO_2 removal to meet Federal NSPS - 78.1% SO_2 removal.

Costs have been developed from the estimates for the LG&E system with appropriate adjustments to the system design and cost factors that are specific to the Cane Run Plant. In order to provide a common basis for comparison of system costs, some of the assumptions used by McGlamery (1975)* in prior studies of the economics of SO_2 scrubbing have been incorporated in the capital and operating cost estimates. The basis for the costs is coal containing 3.5% sulfur and 0.1% chloride and having a heating value of 12,000 Btu/lb.

There are four basic differences between the cost basis for the generalized plants and the LG&E system:

(1) The generalized plants utilize slaked quicklime rather than carbide lime and therefore require slaking equipment and dry lime handling/storage facilities.

(2) The boilers and FGD systems have common fans.

(3) The generalized systems are equipped with flue gas bypass for reheat of the scrubbed gas. While overall SO_2 removal efficiencies of 78% or 90% are achieved, removal efficiencies for the flue gas passing through the absorber may be 98% or higher. In cases where the overall SO_2 removal is 78.1% (to comply with NSPS), the amount of flue gas bypassed is 20% and the amount of reheat provided is 35°F. In the case where overall SO_2 removal is 90%, the amount of flue gas bypassed is 8% and the reheat provided is 14°F. Although reheat is provided with bypassed gas, absorbers are designed to handle 100% of the boiler flue gas.

(4) For the generalized systems, the cost of land for process equipment and sludge disposal has been included in the capital costs (at $3,000 per acre).

The capital and operating cost estimates for the three general cases are shown in Table IV-2 along with the comparable costs for the LG&E system. Costs are given in 1976 dollars (for a plant located in the Midwest) in order to provide a basis for comparison with published data on FGD systems.

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*McGlamery, G.G., et al., "Detailed Cost Estimates for Advanced Effluent Desulfurization Processes," EPA-600/2-75-006, January 1975.

TABLE IV-2

SUMMARY OF ESTIMATED CAPITAL AND OPERATING COSTS (1976 DOLLARS)

Boiler Location	Cane Run Unit No. 6	Generalized		
Boiler Size (megawatts)	300	500	500	1,000
New or Existing	Existing	New	New	New
Basis:				
Operating Hours	5,256	7,000	7,000	7,000
S in Coal, %	5	3.5	3.5	3.5
SO_2 Removal Efficiency	94.2	78.1 (NSPS)	90	78.1 (NSPS)
Sludge Disposal	Onsite after treatment	Onsite	Onsite	Onsite
Reheat, °F	0[a]-50	35	35	14
Capital Investment, $/kw	53.2[b]	50.2	52.8	43.7
Annual Cost, ¢/10^6 Btu	23.0[a]-26.7	32.9	35.8	30.3
Mills, kwh	2.29[a]-2.65	2.96	3.22	2.73

[a]Reheating of the wet gas (from scrubbers) is not in operation (savings in the reheat energy only).

[b]The capital investment for Cane Run Unit No. 6 is equivalent to $53.2/kw based upon gross peak load (300 Mw) and $57.5/kw based upon gross net load (277 Mw).

The major equipment costs for the generalized cases were, for the most part, obtained by adjusting the costs for the LG&E system using standard cost factoring techniques. Due to the modular nature of the system, the sizes of the various pieces of equipment do not differ appreciably from those in the LG&E system, so the use of cost factors results in minimal error. Direct erection labor costs were assumed to be 50% of the direct materials costs, and project indirect costs (engineering, field expenses, fees and contingency) were assumed to be proportional to total direct costs (35% for the 500 megawatt cases and 33% for the 1,000 megawatt system). Higher relative costs for erection labor were used for the generalized systems that were used in estimates for LG&E because of the low cost of construction labor at the Cane Run Plant.

The total capital costs for the generalized 500 megawatt systems are equivalent to about $50/kilowatt (78% SO_2 removal) and $53/kilowatt (90% SO_2 removal). These are about the same as for the Cane Run demonstration system because the savings resulting from economies of scale, absence of reheater, and the lower sulfur content of the coal are roughly offset by the higher costs for erection labor, the need for lime storage and slaking facilities, and the inclusion of the land costs.

The capital cost for the 1,000 megawatt system is estimated at about $44 per kilowatt. This roughly corresponds to a 13% cost savings over the equivalent 500 megawatt system, due to economies of scale.

The operating costs for the generalized systems range from about 30¢ per million Btu to about 36¢ per million Btu (about 2.7 to 3.2 mills/kwh). These cost estimates are about 50% higher than those for the LG&E system, primarily because of the higher costs of lime ($40/ton for commercial quicklime vs. $13.30/ton for carbide lime) and the higher overhead and capital charge factors used.